PB-239 156

ELECTROCHEMICAL CARBON REGENERATION

Paul H. Owen, et al

Environics, Incorporated

Prepared for:

Environmental Protection Agency

June 1972

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SELECTED WATER RESOURCES ABSTRACTS
INPUT TRANSACTION FORM

1. Report No. | 2. | 3. Accession No. PB 239 156

4. Title
ELECTROCHEMICAL CARBON REGENERATION,

5. Report Date
6.
8. Performing Organization Report No.

7. Author(s)
Owen, Paul H, and Barry, John P.

10. Project No.
17020 HAL

9. Organization
ENVIRONICS, INC.
7424 Lorge Circle
Huntington Beach, CA 92647

11. Contract/Grant No.
68-01-0040

13. Type of Report and Period Covered

12. Sponsoring Organization

15. Supplementary Notes

16. Abstract

Electrochemical regeneration of granulated activated carbon spent in the adsorption treatment of municipal secondary effluent was investigated. The adsorption capabilities between electrochemically regenerated carbon and virgin carbon were compared by their respective ability to remove soluble COD organics from sand-filtered secondary effluents during simultaneous dynamic exhaustion.

Electrochemical regeneration was able to restore the working capacity to a level of 42 to 61% of the corresponding virgin capacity over the range of experimental conditions investigated. The 61% regeneration electrochemically obtained is calculated as being 77% of the average obtained by the use of thermal regeneration. The average thermally regenerated capability is known to average approximately 79% of that of virgin carbon.

The electrolysis cell requirement corresponding to the 61% regeneration was measured as 15.3 watthours per gram of COD adsorption capacity regenerated on a spent carbon bed. Based upon the performance with the 61% regeneration, the calculated cost for electrical power at one cent per kilowatt hour to regenerate and rinse the carbon spent in treating 1,000 gallons of Lake Tahoe-type wastewater is 0.66 cents. This cost is equivalent to 3.2 cents per pound of carbon at a dosage of 207 pounds of carbon per million gallons of effluent.

(Owen, Environics)

17a. Descriptors

*Activated Carbon, *Adsorption, *Chemical Oxygen Demand, *Electrolysis, Tertiary Treatment, Waste Water Treatment, Water Purification, Sewage Treatment, Oxidation, Electrochemistry, Electrodes

17b. Identifiers

Regeneration

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Springfield, VA. 22151

PRICES SUBJECT TO CHANGE

17c. COWRR Field & Group 05D

18. Availability

19. Security Class. (Report)
20. Security Class. (Page)
21. No. of Pages 47
22. Price 3.75

Send To:
WATER RESOURCES SCIENTIFIC INFORMATION CENTER
U.S. DEPARTMENT OF THE INTERIOR
WASHINGTON, D. C. 20240

Abstractor | Institution

EPA 670/2-75-028
June 1972

ELECTROCHEMICAL CARBON REGENERATION

by

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Project No. 17020 HAL
Contract No. 68-01-0040

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OFFICE OF RESEARCH AND MONITORING
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

EPA Review Notice

This report has been reviewed by the Environmental Protection Agency and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

ABSTRACT

A process of electrochemical regeneration of granulated activated carbon spent in the adsorption treatment of municipal secondary effluent was experimentally investigated. The adsorption capabilities between electrochemically regenerated carbon and virgin carbon were compared by their respective ability to remove soluble COD organics from sand-filtered secondary effluents during simultaneous dynamic exhaustion at the Irvine Waste Water Treatment Plant.

Data indicate that electrochemical regeneration was able to restore the working capacity to a level of 42 to 61 percent of the corresponding virgin capacity, over the range of experimental conditions investigated. The 61 percent regeneration electrochemically obtained is calculated as being 77 percent of the average obtained by the use of thermal regeneration.

The electrolysis cell energy requirement corresponding to the 61 percent regeneration was measured as 15.3 watt hours per gram of COD adsorption capacity regenerated. Based on this regeneration, the calculated cost for electrical power, at one cent per kilowatt hour, to regenerate and rinse the carbon spent in treating 1,000 gallons of Lake Tahoe-type wastewater is 0.66 cents. This cost is equivalent to 3.2 cents per pound of carbon at a dosage of 207 pounds of carbon per million gallons of effluent.

Cost analyses show that electrochemical regeneration is much less costly than thermal regeneration, particularly when labor cost is considered.

This report was submitted by Environics, Inc., 7424 Lorge Circle, Huntington Beach, California 92647, in fulfillment of Project No. 17020 HAL, Contract No. 68-01-0040 under the sponsorship of the Environmental Protection Agency.

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SECTION 1

CONCLUSIONS

1. Electrochemical regeneration of granular activated carbon spent in the adsorption treatment of municipal secondary effluent can restore the working capacity for removing soluble COD organics to 61 percent of the working capacity of virgin carbon.

2. Cost analyses (Section VI) show that electrochemical regeneration is potentially a much less costly method of regenerating carbon than thermal regeneration. For example, operating costs (exclusive of labor), at 61% regeneration and 1.2 cents per Kw hour, are 80% of operating costs for thermal regeneration. Labor costs are expected to be much less for electrochemical than thermal regeneration. Similarly, if the process can be optimized such that 80% regeneration can be obtained, electrochemical regeneration would be only 60% as costly as thermal.

3. Electrolysis cell energy requirement for the 61 percent regeneration was measured as 15.3 watt hours per gram of COD adsorption capacity regenerated on the spent carbon.

4. The 61 percent of COD capacity electrochemically regenerated is approximately 77% as good as the average thermal regenerated capacity (79%).

5. Based on the performance obtained with the 61% regeneration, the calculated cost for electrical power to regenerate and rinse carbon spent in treating 1000 gallons of Lake Tahoe-type wastewater is 0.66 cents, which is equivalent to 3.2 cents per pound of carbon at a dosage rate of 207 pounds of carbon per million gallons.

6. Electrochemical regeneration from a process viewpoint offers several advantages over the thermal regeneration process, including its suitability for use in small wastewater treatment plants. The process can be automated to require very little labor and much less equipment, since there is no transportation of carbon and fewer operational steps.

7. Because of the lengthy (approximately 60 days) turn-around time in the exhaustion, regeneration, exhaustion cycle, and the relatively short duration of the present contract, it was not possible to optimize the system in terms of all the parameters involved.

SECTION II

RECOMMENDATIONS

As a result of the regeneration capability and cost advantages of electrochemical regeneration demonstrated in this work, the following recommendations are made:

1. Because of the lengthy turn-around time of exhaustion, regeneration, exhaustion cycle, and the relatively short duration of the present contract it is recommended that further work be done in the laboratory using short duration isotherms to optimize the process, following which the optimized parameters will be verified at the Irvine Waste Water Treatment Plant. Should the results appear attractive from the standpoint of cost and other parameters, demonstration on a pilot scale is recommended.

2. Continuation of the work begun in the present study to determine the effects of various combinations of operating parameters such as: electrode materials; brine compositions; cell potentials; cell current; and regeneration obtained in this project is recommended. This should be done on a small scale in the laboratory using 18-hour saturation adsorption isotherms of COD organics from sand filtered secondary effluent. Included in this laboratory study would be construction of a small electrochemical regeneration apparatus with provision made for easy installation and removal of cell electrodes of various materials.

3. If a lead dioxide anode is found feasible, addition of the measurement of the lead content of the carbon bed rinses is recommended. If another electrode material is used, it is recommended that there be an addition of the measurement of a parameter indicative of possible contamination introduced into the rinses by this new material. This type of measurement would also be recommended in case of possible toxic substance produced at the cathode.

4. Reconstruction of the electrochemical regeneration apparatus described in this report to conform to the optimum operating efficiency specifications as determined in the laboratory study is recommended.

SECTION III

INTRODUCTION

Determination of the chemical oxygen demand (COD) provides a measure of the oxygen equivalent of that portion of the organic matter in a sample that can be oxidized by a strong chemical oxidant. It is an important, rapidly measured parameter for control of waste treatment plants as well as other uses. Removal of dissolved COD exerting material is one step toward the purification of treatment plant effluents.

An early study[1] of direct electrolytic treatment of municipal wastewater revealed that dissolved COD organics could apparently be removed from the aqueous solution by electrochemical oxidation at the anode when suitable electrodes were used. However, it was also concluded from this study that the cost depended upon the electrical resistance of the wastewater and the completeness of the COD removal treatment, and the electrochemical method was not considered economically practical for wastewater renovation where waters of relatively low conductivity were to be processed. A later investigation[2], also employing direct electrolytic treatment, showed that dissolved organics could in fact be removed from an aqueous solution by an electrochemical wet-oxidation process, and that they were converted into CO_2, N_2 and H_2O. The investigation further revealed that electric energy costs were only of the order of 20 watthours per gram of COD removed with a wastewater (human urine) of approximately 18,000 μmhos/cm conductivity, which compared remarkably with the cost of 720 watthours per gram of COD removed, as obtained in the early study where the conductivity was only about 500 μmhos/cm. The electrochemical method for removing dissolved organics from wastewater can be economically practical if the electrical resistance of the electrolyte can be made low enough.

An extension of the later investigation showed that a low conductivity wastewater (normal secondary effluent) could be renovated with the same low electric power cost by the electrochemical method when an indirect electrolytic treatment was employed. In this case, activated carbon was first used to remove and concentrate the dissolved COD organics from the wastewater by adsorption, which concentrates the COD from about 10,000 volumes into one volume. The spent activated carbon (drained of the wastewater) was then added to an electrolyte of low electrical resistance and given direct electrolytic treatment. The electrochemical treatment apparently removed the adsorbed COD organics by a wet-oxidation process, which in turn regenerated the carbon's capability for again removing dissolved COD organics from the same type of wastewater. In addition to regenerating the carbon, this investigation also indicated that very low electric power cost (of the order of 5 watthours per gram of COD capacity

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regenerated) may only be required, which would make this process competitive with thermal regeneration. Carbon physical loss was only of the order of 2 to 3 percent, which would reduce carbon regeneration makeup considerably. Because regeneration can occur without removing and transporting the carbon from its adsorption column by means of a simple, completely automated process requiring minimal labor, the electrochemical regeneration process would be applicable to relatively small capacity waste treatment plants. The problem of bacterial growth within carbon beds, fouling and clogging could be controlled because the electrochemical process generates sterilizing oxidants, and the problems of air pollution associated with the thermal regeneration combustion-oxidation of the COD organics was eliminated because the products of the room temperature wet-oxidation electrochemical process are CO_2, N_2, and H_2O.

All of the results discussed above for the electrochemical regeneration process for carbon were originally obtained from small scale laboratory studies using small platinum electrodes with gram amounts of carbon, and comparing carbon performance from 18-hour saturation adsorption isotherms. To further the state-of-the-art, similar favorable results would have to be obtained with a larger scale field test study, to show that the electrochemical regeneration process was attractive technically, and could be economically practical. And most importantly, these field test results would have to be obtained with the use of inexpensive nonplatinum electrodes, and the regenerated carbon performance would have to be measured from actual carbon bed working capacities. Accordingly, the objective of the present work under this contract is to experimentally evaluate the process of electrochemical regeneration of granular activated carbon field spent in the purification of municipal wastewater. The measurement or determination of the following items as specific experimental program objectives is required for evaluation of the electrochemical regeneration process:

1. the multiple regeneration efficiency;
2. regeneration energy and time requirements;
3. the performance and useful life of granular activated carbon reactivated by this process in terms of soluble-COD removing working capacity and carbon makeup regeneration losses for comparison with other means of reactivation;
4. the advantages and disadvantages associated with this process in terms of simplicities and complexities, waste products and their ultimate disposal; and
5. possible problems such as equipment corrosion or water recontamination.

SECTION IV

TEST APPARATUS AND PROCEDURES

General

The test apparatus consists of an adsorption system and a regeneration system. The adsorption system is installed at the Irvine Wastewater Treatment Plant and the regeneration system is installed at the Environics Laboratory. The test procedure began with the placement of virgin activated carbon in glass pipes which served as contactors to remove soluble COD organics from secondary effluent of the Irvine Wastewater Treatment Plant. The contactors were then installed on the adsorption system. Following the exhaustion of a virgin carbon bed, it was returned to the laboratory, subjected to an electrochemical regeneration procedure using a low resistance brine, and then returned to the wastewater treatment plant for re-exhaustion. The exhaustion, regeneration, exhaustion cycle was repeated as many times as possible, depending upon the resultant performance. For each regenerated bed re-exhausted a contactor containing a virgin carbon bed was simultaneously exhausted in order to determine the effectiveness of the regeneration.

Samples of the effluent of each column and of the column feed were taken hourly for approximately 30 days. Thus, three composite samples were obtained of the secondary effluent. From analysis of these samples the weight of adsorbed COD exerting organic material on a bed was determined and by comparison of this weight of COD on the regenerated and the virgin bed the effectiveness of the regeneration was calculated.

Carbon Columns

The contactor columns were 2 in. ID by 4-ft long pyrex glass pipes, containing 600 g of granular activated carbon which produced carbon beds that approximately averaged 63 cm in length. Immediately above the 2 in. glass pipe, connected by a bell reducer, was a 3 in. ID by 2 ft section of glass pipe into which the carbon bed could expand during the regeneration procedure. (The flow during regeneration being the reverse of that during exhaustion.) The carbon used in all columns was Filtrasorb 400 (Calgon Corporation), 12 x 40 mesh. Manufacturer's data indicated a mean particle diameter of 1 mm, Iodine No of 1100, and Molasses No of 200.

Adsorption System

A system was constructed to field test the adsorption performance of the carbon beds that were regenerated by electrochemical treatment. This system, which

is shown in Figure 1, was installed at the water reclamation plant of the Irvine Ranch Water District at Irvine, California. The adsorption system consisted of space for up to three carbon columns; two sand filters, a 24-hour composite sampling system, connecting PVC plumbing, and a frame stand for support of the contactor columns. The one-stage downflow hydraulic application rate was 4 gpm per square foot for a superficial contact time of 3.7 minutes. A refrigerated composite liquid sampler sampled the common feed and the output of each column. The sand filter system consisted of two 20-40 mesh sea sand filter beds connected in series to provide the common feed to the carbon. The downflow hydraulic application rate was 7 gpm per square foot. These filters were each plumbed to allow individual back-flushing as required and provisions for individually backflushing each carbon column were also made. Both the carbon columns and the sand filters were covered with aluminum foil to inhibit the reproduction of organisms capable of photosynthetic activity.

Regeneration System

The regeneration system, housed in the Environics Laboratory, consisted of an electrochemical gas-liquid separation cell wired to a DC power supply; a pump and associated plumbing and valving to recirculate brine through the cell and exhausted contactor column; a metal stand to support the contactors during the regeneration process; and a gas chromatograph to monitor the gases produced in the regeneration cell.

The case of the electrochemical cell had outside dimensions of 9 inches deep by 13 inches wide by 48-1/4 inches high, and was constructed of machined 1/2 inch Lucite, as were the top and bottom flanges. The entire cell was mounted on the base of a hand operated crane to facilitate easy removal of the top flange and the attached electrode assembly for cleaning of the cell components (Figure 2).

The entire electrode assembly was mounted on the top flange with titanium steel bolts, some of which served as the external terminals for connection of the power supply and recorder. The electrode assembly of the cell was of a sandwich configuration with the lead dioxide anode placed between the two-piece, Type 347 stainless steel cathode. The two pieces of 1/32 inch sheet stainless steel were mounted on 1/2 inch Lucite and provision was made for adjustment of the distance between each of these and the anode.

The configuration of the electrode assembly caused the cell to be divided into two parts; one, an active electrolysis area between the electrodes, and two, a brine storage reservoir between the cell case and the Lucite cathode supports. Most leakage between the two areas was prevented by part of the weight of the electrode assembly borne on a silicone rubber gasket on the bottom flange. PVC plumbing (used throughout the entire system) was connected, through the bottom flange, to each area to carry the regenerative brine into and out of the cell.

Part of the weight of the electrode assembly was also borne by a silicone rubber gasket between the top flange and the cell case. The purpose of this gasket was to prevent the escape of gases produced during regeneration. These gases could then be examined by gas-chromatography, (Figure 3).

Gas Chromatography

A line of polypropylene tubing connected to the top flange of the cell lead to a Perkin-Elmer Model 990 dual column gas chromatograph equipped with a thermal conductivity detector. Both columns were 1/8 in. by 12 ft stainless steel packed with 80-100 mesh Chromasorb 102. The injector and manifold of the GC were operated at ambient temperature, the columns at approximately 100°C, and the detector at approximately 150°C. The carrier gas was helium at a flow rate of 30 ml/min.

Power Supply

Power for the regeneration process was supplied by a Udylite, Model L-300 21PL-3C012, three phase, 230 volt, 12.2 ampere rectifier which had a maximum output of 300 amperes at 12 volts DC.

Provision was made for continuous recording of time versus power input to the cell by use of an L and N dual channel Speedomax Recorder. The applied cell voltage was recorded on one channel of the recorder and the amperage on the second channel by use of a 500 ampere, 50 millivolt, precision shunt between the positive pole of the power supply and anode of the cell.

Water Analysis

The composite samples taken at the wastewater treatment plant were removed to the laboratory every 18 to 96 hours for immediate analysis. At the same time that the composite samples were retrieved, as well as at the start of the initial composite sampling period, a grab sample was taken of the raw secondary effluent (sand filter input). Portions of the composite samples were filtered through a HA 0.45μ Millipore filter and the dissolved COD of each of these samples determined by the acid dichromate reflux procedures given in Standard Methods.[3] The total COD (unfiltered through 0.45μ filter) of the sand filter output and the sand filter input were also determined. The conductivity, pH, and turbidity of an unfiltered portion of each of the composite samples and of the grab sample were measured, respectively, by a YSI Model 31 Conductivity Bridge, a Sargent-Welch Model LSX pH meter, and a Hach Model 2100 Turbidimeter. These latter three parameters were used as probable indicators of possible recontamination introduced into the treated wastewater from the carbon subjected to the regeneration process.

Calculation of Adsorbed COD

The determination of adsorbed dissolved COD exerting material, in grams, on each carbon bed is found by the formula:

$$\sum_{t=0}^{t=n} (\bar{X}e)_t \, (T)_t \, (\Delta COD)_t \times 10^{-3}$$

for each composite sample interval where:
$\bar{X}e$ is the mean effluent flow rate through the carbon bed in liters per minute,
T is the length of the interval in minutes,
n is the total number of intervals, and
ΔCOD is the dissolved COD in milligrams per liter of the carbon column feed less that of the carbon column effluent.

Regeneration Brines

Brines of 1.48 to 6.64 percent salt content (sodium sulfate or sodium chloride or both) were made and stored in 28 liter polyethylene tanks, from which they could be pumped into the recirculating regeneration system by proper valving. The exact composition; pH; conductivities; dissolved COD after use; and other notes concerning final color, appearance, odor, and regeneration temperature range of the various brines before and after use in a regeneration are discussed in Section V. Chloride content of the brines, as well as that of the rinses was determined by use of a Corning Chloride ion electrode (Catalogue Number 476126) in conjunction with the Sargent-Welch pH meter.

Recirculating Pump

The pump to recirculate the brine was a Little Giant Model 4MD (Catalogue Number 582005) powered by a 1/2 HP, 3100 RPM, 180 W, 115 volt, Model ABISOC5N# motor.

Regeneration Procedure

A regeneration was begun by placing an exhausted carbon column, which had previously been rinsed 6 to 9 times with one to two liters of tap water to remove as much as possible of the solid suspended material, on the regeneration stand and connecting the plumbing to the top and bottom fittings of the carbon contactor. Brine was then pumped into the system and allowed to begin to recirculate through the system.

Regeneration brine was first pumped from a storage tank; then through a valved

bypass, around the carbon contactor, and then through plumbing connected to the bottom flange of the cell into the active electrolysis area. After this active area was completely filled, the brine would spill over the Lucite cathode supports into the reservoir area of the cell. After all of the brine was pumped out of the storage tank and into the recirculating system the storage tank was valved off and the flowmeter needle valve opened. From the reservoir, the brine flowed out of the cell through plumbing, again connected to the bottom flange of the cell; through the pump; through the combination flowmeter and metering valve mounted on the contactor support stand; upward through the exhausted contactor, thus expanding the carbon bed; and then back into the regenerative area of the cell (Figure 4). A valved bypass around the flowmeter and carbon contactor allowed the pump to operate at all times at full flow, thus preventing unnecessary wear. Initially, in a regeneration, the flow rate through a column was 1000 ml/min. This flow rate was increased in 500 ml/min increments every 15 minutes to a maximum of 2000 ml/min. (2,500 ml/min in Regeneration No 1.) The purpose of this gradual increase in the flow rate was to prevent the escape through the unscreened plumbing on the top flange of the contactor of any carbon into the remainder of the system.

The upflow hydraulic application rate (HAR) for the carbon bolumns at the above maximum flow rate was 24.3 GPM per square foot. (For carbon bed R1 only, HAR was 31.5 GPM per square foot.)

At approximately the same time that the 1000 ml/min flow rate through the column was established, power was applied to the cell and the recorder activated. Small adjustments were made in the cell voltage every 15 minutes for the first 0.5 to one hour of regeneration time and then as needed, in order to maintain a constant cell voltage ±0.025V.

At various times during a regeneration, the gases produced were analyzed by gas-chromatography.

Once the desired amount of energy had been applied, the rectifier was switched off, the recirculating pump stopped, and the cell, column and plumbing drained. The carbon bed contained in the contactor was then rinsed with the desired number of changes of water to remove brine adhering to the carbon. These rinses were added to and drained from the contactor through a valved tap entering the system physically below the contactor (Figure 4). From four to twelve rinses were performed on any one regenerated carbon bed. The type of water used; duration of time in a bed and time allowed for draining; pH; conductivity; dissolved COD; and comments concerning color, appearance and contaminants of each rinse, are discussed in Section V.

Nomenclature

The following simplified system has been adopted in this report for identifying the various carbon beds used in the program. Regenerated carbon beds are designated R1 through R7 to correspond to the seven regenerations that were performed. Virgin carbon beds used as comparison to the regenerated beds during the subsequent adsorption tests are designated V1 through V6. (Regenerated beds R6 and R7 were both compared to the same virgin bed, V6).

Brines used in the seven regenerations are designated B1 through B7, respectively.

SECTION V

RESULTS AND DISCUSSION

During this project seven regenerations of spent carbon beds were performed. The results of this work are summarized in the following tables and figures. The comparative performance of each carbon bed after electrochemical regeneration is presented in Table 1; the actual wastewater adsorption performance of each carbon bed, virgin and regenerated, is presented in Table 2; the power consumption for each regeneration is given in Table 3; and the regeneration apparatus operating parameters used in each regeneration are given in Table 4.

The success of the various regeneration procedures is evaluated by the calculation of the parameter "carbon regeneration efficiency" (CRE), defined as:

$$CRE = \left(\frac{\text{COD loading on regenerated bed at breakthrough}}{\text{COD loading on comparison virgin bed at breakthrough}}\right) \quad (100\%)$$

The details of each regeneration are discussed individually, below.

Regeneration of Carbon Bed R1

Prior to the first regeneration, a virgin carbon bed was exhausted by treating 10,347 liters of wastewater at the Irvine plant. The bed was on stream for 552 hours, and was removed for regeneration when a composite sample of its effluent had a soluble COD value 85.6 percent of that of the untreated sand filter effluent. At this time the COD loading on the bed was calculated to be 107g.

This bed, designated R1, was subjected to a 24.0 hour regeneration procedure, which was conducted at a constant cell potential of 4.00 volts and at a current of 41.0 amperes. The energy input for this regeneration was 3940 watt hours, i.e. 36.6 watt hours per gram of COD organic adsorbed (Table 3). The R1 regeneration was performed using a 1.48 percent sodium chloride brine (B1). This brine was composed of 14.9 g/l NaCl in RO water, and had a pH of 7.55 and conductivity of 25,500μmho/cm. This brine developed a dark red-brown color, was slightly opaque, and had a strong chlorine odor at the end of the regenerative procedure.

The R1 regeneration was conducted with an electrode spacing of 0.5 in., and used 40.0 liters of brine at an upflow hydraulic application rate (HAR) of 31.5 gallons per minute (GPM) per square foot of bed area. This HAR produced an

CARBON BED ELECTROCHEMICAL REGENERATION RESULTS

Regenerated Bed No.	Comparison Virgin Bed No.	Virgin Adsorption (%)	Final Breakthrough (%)	Regenerated Adsorption (g)	Final Breakthrough (%)	Energy Input Per Gram of COD Capacity Regenerated (watt hours)	Carbon Regeneration Efficiency (%)
R1	V1	44.0	57.4	3.27	97.6	1204	7.44
R2	V2	65.1	86.4	39.80	87.6	15.3	61.10
R3	V3	77.5	86.8	32.70	99.6	36.3	42.20
R4	V4	58.1	54.3	25.10	81.2	40.8	43.30
R5	V5	49.2	82.7	21.50	86.9	17.0	43.60
R6	V6	98.0	64.0	36.60	88.8	6.28	37.40
R7	V6	98.0	64.0	46.10	92.5	3.42	47.10

Table 1.

CARBON BED WASTEWATER ADSORPTION PERFORMANCE

Carbon Bed No.	Volume of Wastewater Treated (liters)	Time On Stream (hours)	Last Composite Sample Breakthrough (%)	Dissolved COD Adsorbed (g)	COD Loading (g COD / 100g carbon)
R1	4,527	236	97.6	3.27	0.545
V1	4,445	236	57.4	44.00	7.33
R2	7,920	575	87.6	39.80	6.63
V2	8,040	575	86.4	65.10	10.80
R3	8,712	624	99.6	32.70	5.45
V3	9,350	624	86.8	77.50	12.90
R4	4,742	296	81.2	25.10	4.19
V4	4,589	296	54.3	58.10	9.68
R5	5,466	344	86.9	21.50	3.58
V5	5,222	344	82.7	49.20	8.21
R6	10,551	704	88.8	36.60	6.10
R7	9,705	663	92.5	46.10	7.68
V6	9,432	704	64.0	98.00	16.30

Table 2

REGENERATION CELL POWER CONSUMPTION

Bed No.	Regen. Duration (hours)	Average Cell Potential (volts)	Average Cell Current (amps)	Average Anode Current Density (ma/cm^2)	Power Input per gram of COD adsorbed (watts)	Energy Input per gram of COD adsorbed (watt hours)
R1	24.00	4.00	41.00	13.70	1.53	36.00
R2	30.50	3.08	6.48	2.16	0.26	8.00
R3	70.00	2.98	5.67	1.89	0.42	29.70
R4	36.00	3.00	9.50	3.17	0.43	15.80
R5	61.00	2.81	2.13	0.71	0.07	4.70
R6	6.25	3.20	12.50	4.16	1.86	11.60
R7	3.75	3.20	13.30	4.44	0.86	3.24

Table 3

CARBON BED REGENERATION CELL PARAMETERS (a)

Regeneration No	Electrode Spacing (inches) and Anode Area (cm^2)	Total Volume of Brine (liters)	Col. Brine Flow Rate (l/min)	Observed Cell Changes and Regeneration Comments
R1	1/2	40.0	2.6	No deterioration of electrodes or cell. Fine brown/black coating found on cathodes.
R2	5/16	40.0	2.0	No deterioration of electrodes or cell. Slight fine white coating found on cathodes.
R3	5/16	26.0	2.0	No deterioration of electrodes or cell. Fine white coating found on cathodes.
R4	5/16	24.0	2.0	No deterioration of electrodes or cell. Fine white coating found on cathodes.
R5	5/16	24.0	2.0	No deterioration of electrodes or cell. Fine white coating found on cathodes.
R6	5/16	21.0	2.0	No deterioration of electrodes or cell. Fine white coating found on cathodes.
R7	5/16	21.0	2.0	No deterioration of electrodes or cell. Fine white coating found on cathodes.

(a) For all regenerations, anode material was lead dioxide, with surface area of 3000 cm^2, and cathode material was type 347 stainless steel.

Table 4

A Photograph of Environics' Activated Carbon Adsorption Test Apparatus at the Irvine Ranch Water District Reclamation Plant

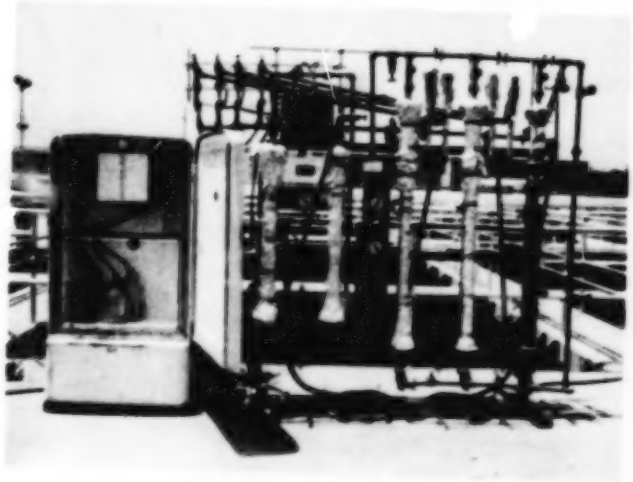


Figure 1

Photograph of Carbon Bed Undergoing Regeneration in Glass Contactor

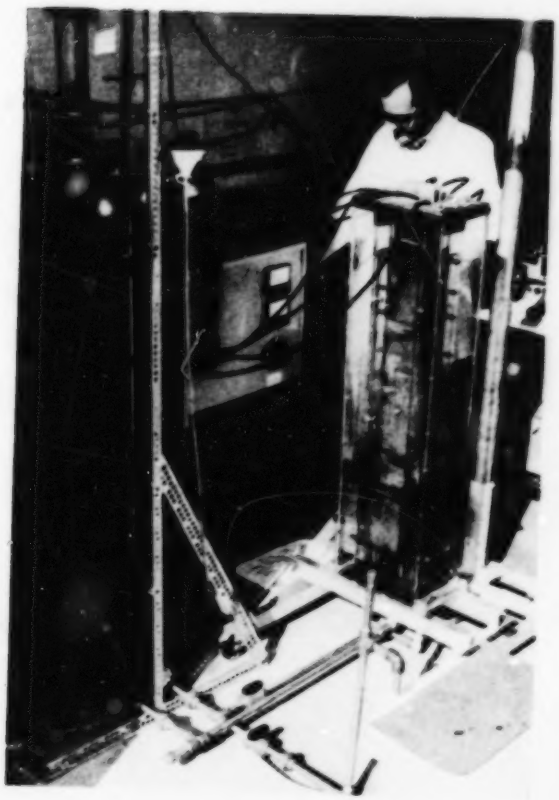


Figure 2

Photograph of Electrolytic Regeneration Cell

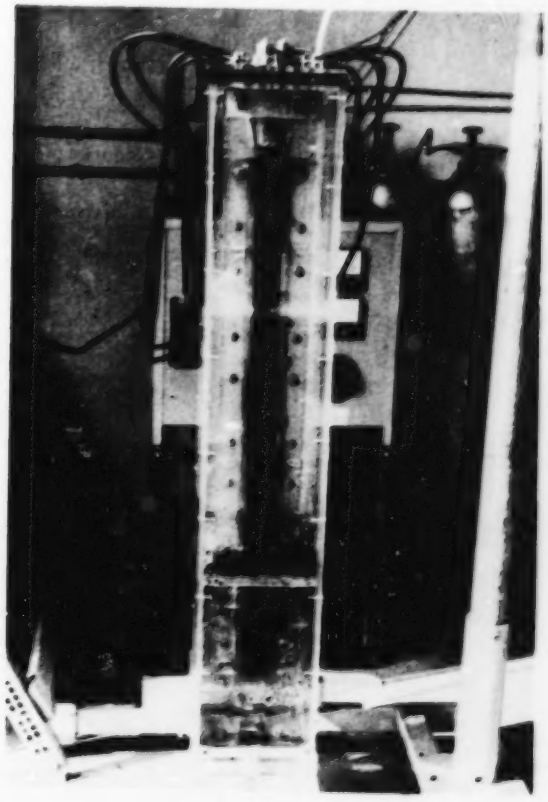


Figure 3

Diagram of Brine Recirculation in Regeneration Apparatus

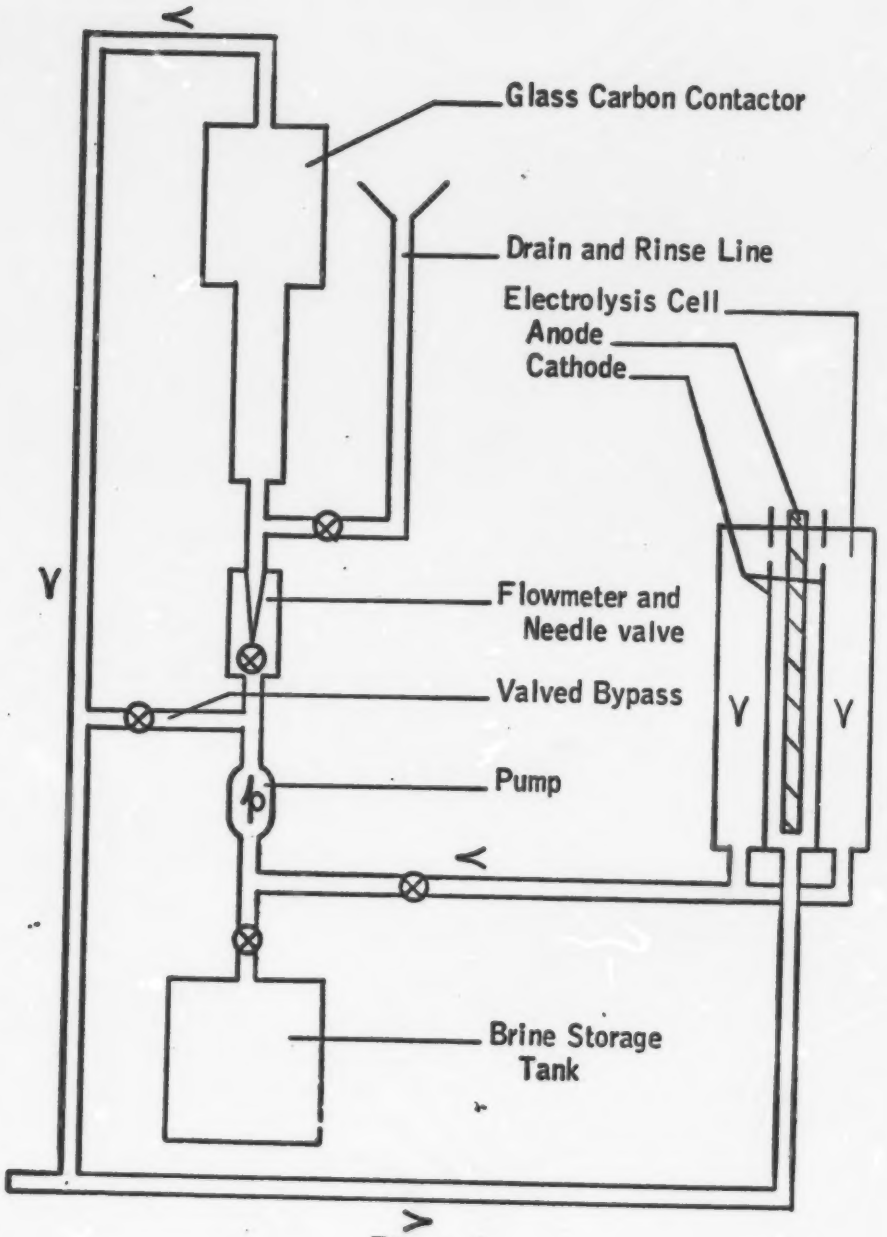
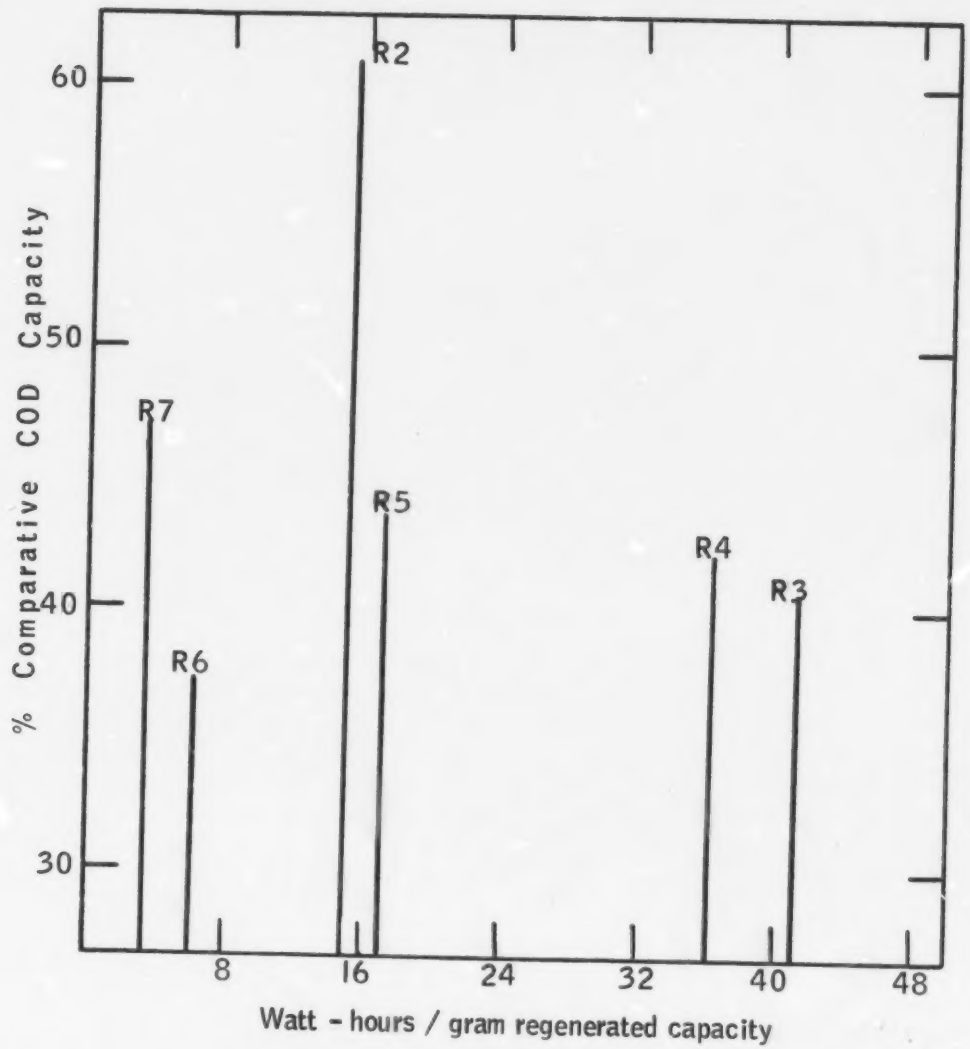
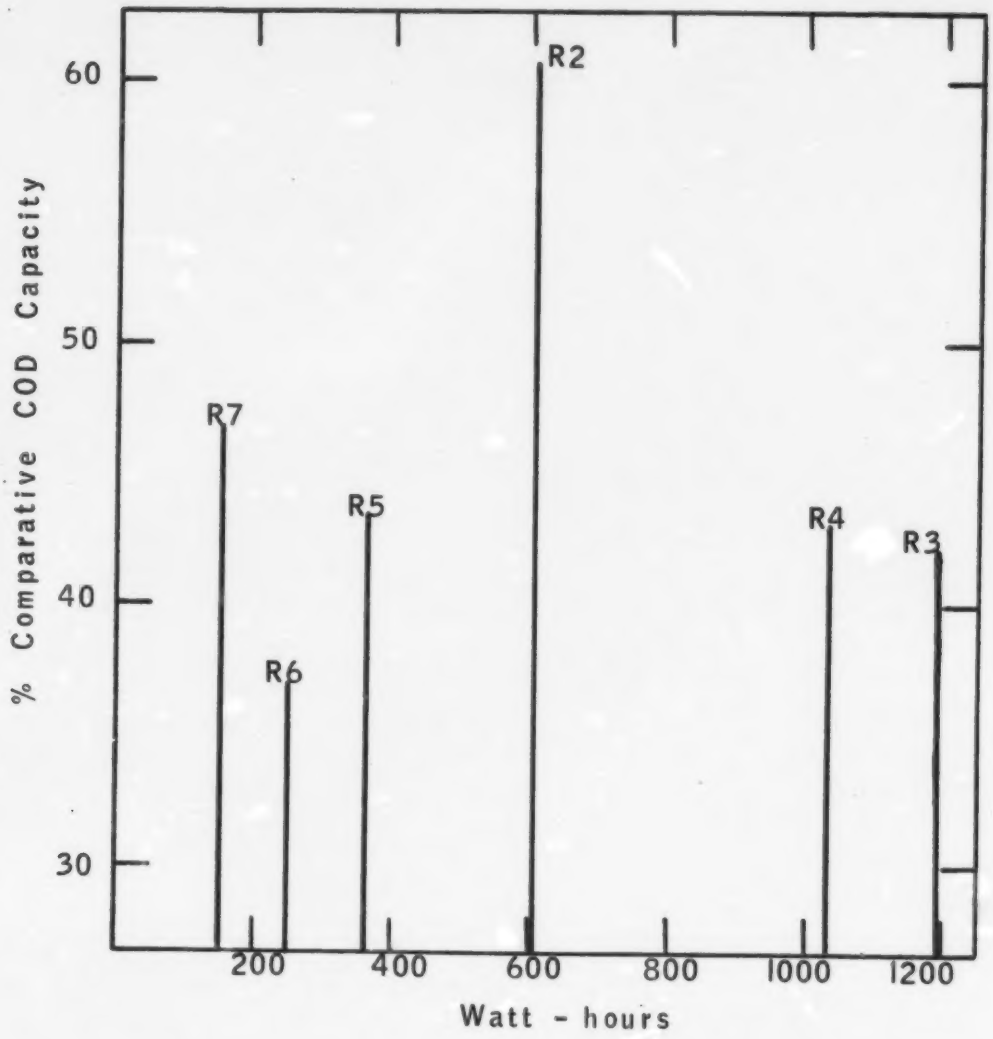


Figure 4



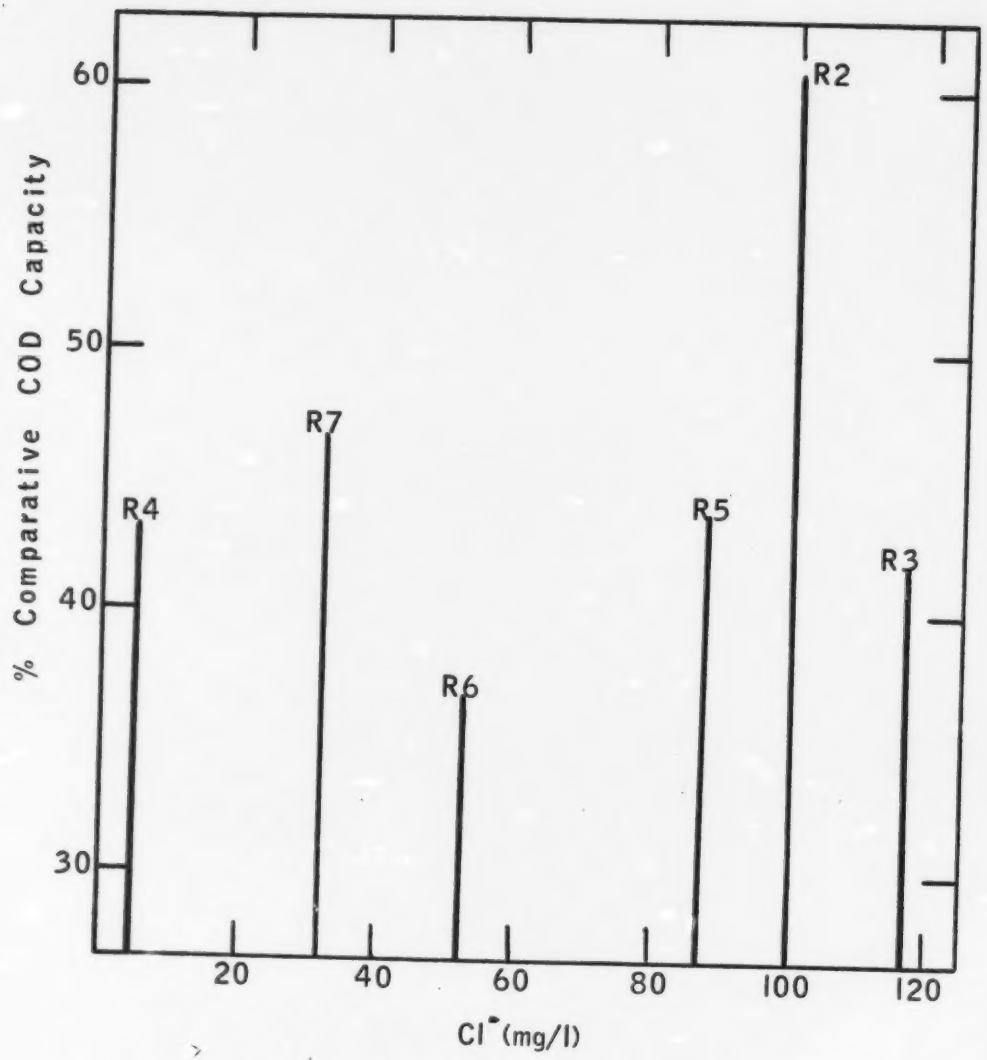
Bar Chart of Energy Input (watt-hours) per Gram of COD Capacity Regenerated and Comparative COD Capacity (%) after Regeneration R2 - R7.

Figure 5.



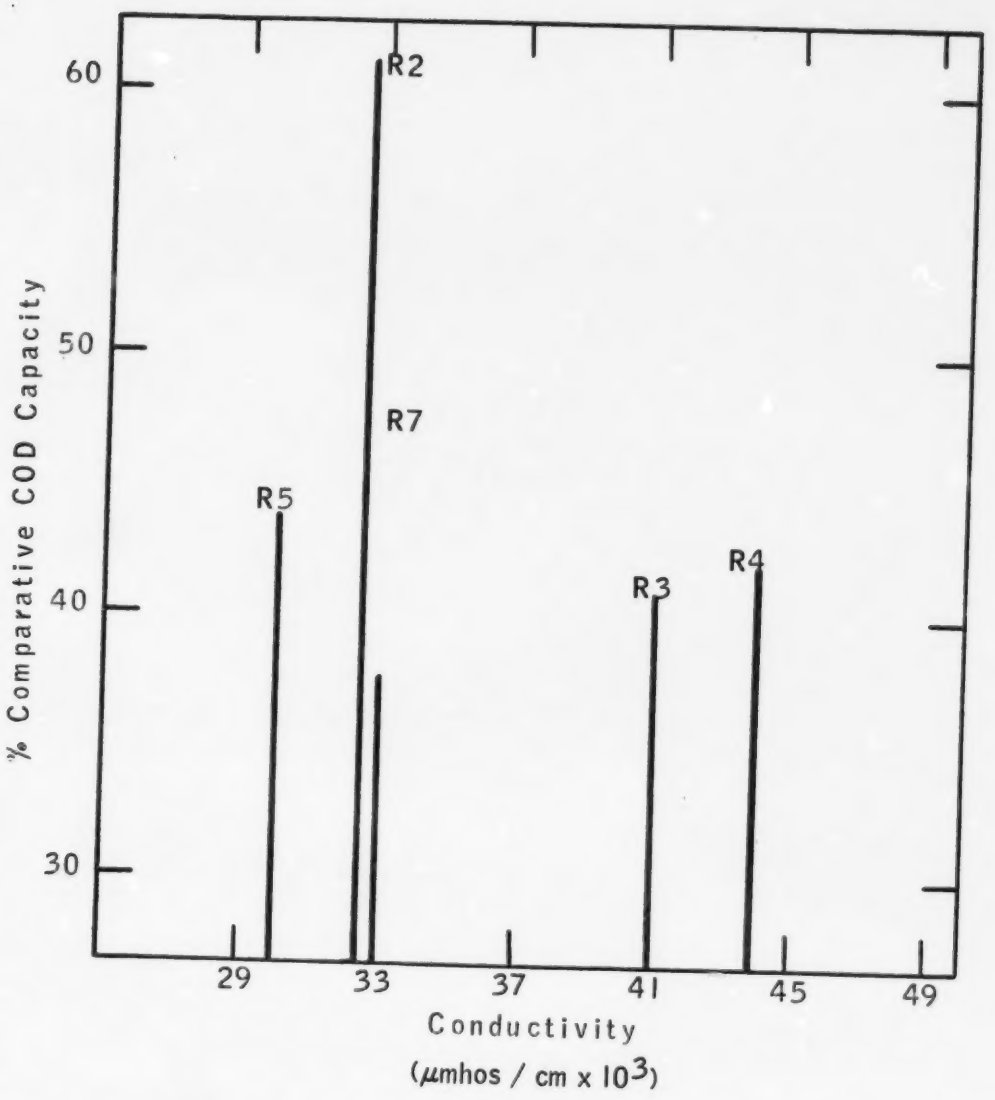
Bar Chart of Total Energy Input (watt-hours) for Regeneration Procedures R2 - R7 and Comparative COD Capacity of % Regenerated Carbon Beds.

Figure 6.



Bar Chart of Chloride Concentration (mg/l) of Brines Used in Regeneration Procedures R2 - R7 and Comparative COD Capacity (%) of Carbon Beds After Regeneration.

Figure 7.



Conductivity of Brines Used in Regeneration Procedure R2 through R7, and COD Capacity as Percentage of Virgin Performance.

Figure 8

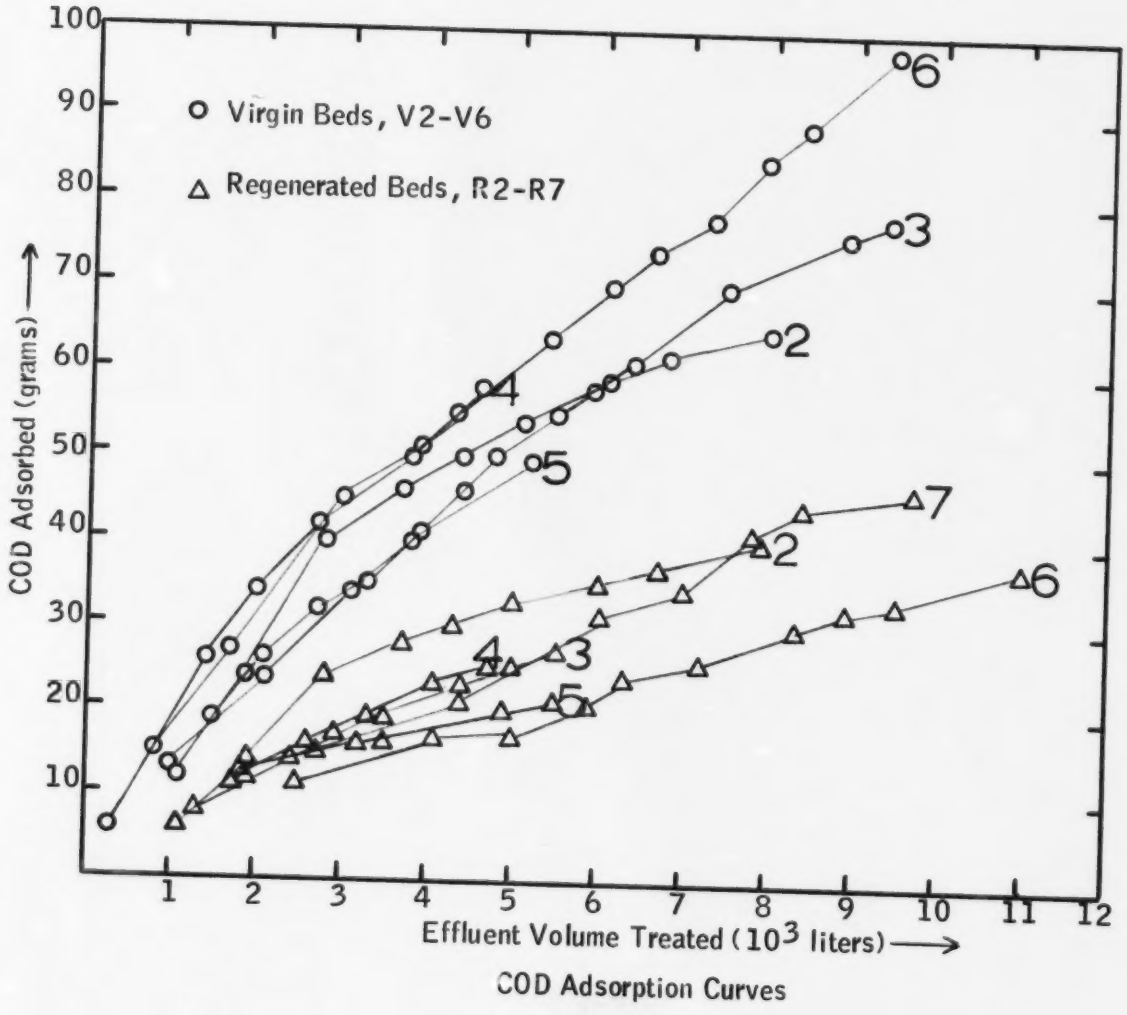

Virgin Beds, V2-V6
Regenerated Beds, R2-R7
COD Adsorbed (grams)
Effluent Volume Treated (10^3 liters)
COD Adsorption Curves

Figure 9

expansion of 117 percent of the downflow bed length. Near the end of the R1 regeneration treatment, the gas evolved from the electrolysis cell was examined by gas chromatography. The approximate composition of the gas was as follows:

48 percent hydrogen
25 percent nitrogen
12 percent combined carbon monoxide and oxygen, and
15 percent carbon dioxide.

These values, as well as those reported from any other regenerations, are only semiquantitative because it was found that the construction of the electrolysis cell did not produce a leak-proof chamber. Major leakage was found around the silicone rubber top flange gasket and around the bolts used to anchor the electrode assembly to the top flange.

Following electrochemical treatment, bed R1 was rinsed in its contactor column, with three changes of reverse-osmosis treated water and once with tap water. The RO treated water had a conductivity of 82 μmho/cm and a pH of 7.80. The tap water had a conductivity of 725 μmho/cm and a pH of 7.55. The rinses remained in the bed from 15 to 30 minutes. Fifteen minutes were allowed for the draining of each change of water before the next rinse was added to the carbon column. All of these rinses had a dark color, and the fourth had a high dissolved COD value (293 mg/l). Final pH of each rinse was $\sim$4.5 and conductivity decreased from 11,500 μmho/cm for the first to 1,420 μmho/cm for the last rinse.

Following the rinse procedure, carbon bed R1 was placed on stream for exhaustion. It ran for 236 hours and treated 4,527liters of wastewater. The last composite sample of the effluent of this column had a dissolved COD value 97.6 percent of the untreated feed wastewater. The bed absorbed only 3.27 grams of COD organics or 0.545 grams per 100 grams of carbon (Table 2).

Comparison virgin bed V1 was also on stream for 236 hours, saw 4445 liters of sand filtered effluent, and had a COD breakthrough 57.4 percent of the untreated effluent. By comparison with bed V1, the regenerated bed R1 had a CRE of only 7.44%.

The poor performance of carbon bed R1 after regeneration was probably due to the detrimental effect of excess chlorine which was generated by the high power input in conjunction with the high chloride content of the regeneration brine. That is, the oxidizing efficiency of the B1 brine was so great at the power applied that it attacked the carbon as well as any adsorbed COD organics. This would account for the final appearance of the brine and of the rinses of the bed. (Due to the obviously poor choice of operating parameters, this

regeneration is not included in Figures 5, 6 and 7.) In order to attempt a more effective regeneration, and not peptize the carbon in a bed while undergoing regeneration, the composition of the brine (B2) for use in the next regeneration was changed. This brine was used to electrochemically regenerate carbon bed R2 in the second regeneration, described below.

Regeneration of Carbon Bed R2

A virgin carbon bed was placed on stream and exhausted by being used to treat 9,878 liters of wastewater. After being on stream 502 hours, it had a final COD breakthrough 93.2 percent of the untreated effluent, and had absorbed 76.0 grams of dissolved COD organics. The carbon column was then returned to the laboratory for regeneration.

Carbon bed R2 was subjected to 30.5 hours of regeneration at a potential of 3.08V and current of 6.48 amps. The input regeneration energy was thus 609 watt hours, or 8.00 watthours per gram of COD adsorbed on the bed (Table 3).

The B2 brine used in R2 contained principally sodium sulfate (35.4 g/l), with only 100 mg/l of sodium chloride. It had a pH of 7.35 and conductivity of 32,500 μmho/cm. As stated, the formulation was based on the results of the poor performance of R1, which was probably peptized due to the over production of chlorine or oxidizers containing chloride.

The R2 regeneration was conducted using an electrode spacing of 5/16 inch and 40.0 liters of brine. The upflow HAR of 24.3 GPM per square foot of carbon bed area produced an expansion of approximately 90 percent of the bed length in the contactor column. This hydraulic application rate was used in all subsequent regenerations (R3 through R6), where it produced approximately the same degree of bed expansion. The same electrode spacing was also used in subsequent regenerations.

The gas evolved from the electrolysis cell during the R2 regeneration was examined by gas chromatography and was found to be composed of the following:

36.6 percent hydrogen
21.3 percent nitrogen
40.8 percent combined carbon monoxide and oxygen, and
1.3 percent carbon dioxide.

The lower percentage carbon dioxide than in the R1 regeneration seems to indicate that a milder oxidation was occurring in this regeneration. This is probably due

to the absence of oxidation of the activated carbon bed itself, in contrast to what probably occurred in R1.

After treatment, this bed was rinsed four times with 1.2 liter volumes of tap water (rather than RO water) to duplicate practical operating conditions. The final pH of each rinse was $\sim$7.3. Final conductivities decreased from 11,100 μmho/cm to 1,220 μmho/cm. Each rinse was clear and colorless, with a trace of carbon fines settled out. From these results it appears that the brine regeneration salts are easily rinsed from a carbon bed and therefore electrochemical regeneration should not cause or contribute to any salt contamination problem.

After regeneration, bed R2 was placed on the wastewater stream and exposed for a continuous period of 24 days to 7,920 liters of wastewater. The last composite sample gave a breakthrough of 87.6 percent when 39.8 grams of COD organics had been adsorbed on the bed.

For the comparison virgin bed, V2, the last composite sample of effluent had a dissolved COD value 86.4 percent of that of the sand filter effluent. This comparison carbon bed saw 8,040 liters of wastewater and adsorbed 65.1 g of COD organics (Table 2). From these figures the carbon regeneration efficiency for R2 was calculated as 61.1%. This regeneration required approximately 16 watthours per gram of COD capacity restored.

Regeneration of Carbon Bed R3

Because of the improvement in performance of carbon bed R2 (61.1% CRE) over that of R1 (7.44% CRE), the carbon bed that had been regenerated and spent in R2 was subjected to a second regeneration, designated R3. This regeneration was performed in an attempt to demonstrate a multiple regeneration and, at the same time, effect a more complete regeneration (greater than 61.1%). To accomplish this more effective regeneration, the carbon bed in R3 was subjected to a regeneration procedure of 70.0 hours duration. The average cell potential for this regeneration was approximately 2.98 volts, and the average cell current was 5.67 amperes. The energy input was 1182 watt hours, or 29.7 watt hours per gram of adsorbed COD organics.

In R3, the B2 brine was again used. However, due to evaporation, the concentration of this brine was increased from 35.4 g/l sodium sulfate and 100 mg/l chloride in 40 liters of water to 49.5 g/l sodium sulfate and 16 mg/l chloride in 26 liters of water. These values are based on conductivity measurements, and therefore are only approximations. The chloride content of the brine was increased to 116 mg/l before R3 was initiated (Table 5). This adjusted brine

is designated B3.

The composition of the gas evolved from the electrolysis cell during the last three hours of the R3 regeneration was as follows:

21.1 percent hydrogen
40.8 percent nitrogen
37.3 percent combined carbon monoxide and oxygen, and
0.84 percent carbon dioxide.

After regeneration, carbon bed R3 was rinsed extensively with twelve 1.2 liter volumes of tap water. The final pH of all twelve rinses was ~ 7. Conductivity decreased from 13,500 μmho/cm to 770 μmho/cm. A very slight tannish color was also noticed for the first time beginning with rinse No 3 and extending through rinse No 10 before fading out. A trace of what appears to be carbon fines settling out was also noticed in all except the first of the twelve rinses.

Following these rinses, carbon bed R3 was placed on stream along with virgin carbon bed V3, and both columns remained on stream for 624 hours. At the end of 624 hours on stream, the 99.6 percent breakthrough level was reached with carbon bed R3, while the V3 bed had a breakthrough of 86.8 percent. At this time bed R3 had adsorbed 32.7 g COD and bed V3 had adsorbed 77.5 g COD. The CRE of R3 was thus 42.2 percent (Table 1). This 42.2 percent regeneration required over 40 watt hours per gram of COD capacity regenerated.

The lower performance of R3 compared to R2 (Table 1) seemed to indicate that only 100mg/l of chloride in the electrolysis brine could be gradually detrimental to the regeneration of activated carbon when using the lead dioxide anode. The word "gradually" must be emphasized because the 100 mg/l chloride level used in the B2 brine for the R2 regeneration had not apparently adversely affected the carbon regeneration.

Regeneration of Carbon Bed R4

With the objective again being to obtain a better initial regeneration (improve on the 61.1 percent performance of R2), regeneration procedure R4 was initiated upon carbon bed V2. This virgin carbon bed had treated 8,040 liters of wastewater, was on stream 575 hours, had a final composite effluent sample with a dissolved COD value 86.4 percent of the untreated wastewater and had adsorbed 65.1 grams of dissolved COD organics (Table 3).

The R4 regeneration used brine B4. This brine was made using distilled water and sodium sulfate (70.4 g/l) only. The chloride content was measured as only 5 mg/l, and pH at 7.17. As possible compensation for the lack of chloride

containing oxidizers the B4 brine was made slightly more conductive (44,000 μmho/cm at 20.6 C) than previously utilized brines.

The R4 regeneration was conducted for 36 hours at an average cell potential of 3.00 volts which, with the increased conductivity of the brine, produced an average cell current of 9.50 amperes and an average anode current density of 3.17 ma/square centimeter. At the higher cell potential and current, a shorter duration of time was necessary to apply approximately the same total energy (1030 watt hours) and energy per gram of adsorbed COD organics (approximately 15 watt hours per gram) as the R3 regeneration (Table 3).

The gases evolved from the electrolysis cell during the R4 and subsequent regenerations were not analysed by gas chromatography. Due to the excessive leakage of gas from the cell, it was decided that these analyses would not be profitable in terms of useful information.

Following regeneration, bed R4 was rinsed with six changes of water. This number of rinses was determined as the optimum number based on conductivity measurement. That is, if six changes of water were each allowed to remain in a bed for 180 minutes and the bed allowed to drain for fifteen minutes between each rinse, the conductivity (1245 μmho/cm) of the last rinse would be approximately the same or slightly less than that of the Irvine plant chlorinated secondary effluent (1500 μmho/cm).

Following rinsing, bed R4 was placed back onstream for 296 hours during which time it treated 4,742 liters of wastewater. After this short length of service, its composite effluent sample had a dissolved COD value 81.2 percent of that of the untreated sand filter effluent. At this point R4 had adsorbed 25.1 grams of COD organics. At this time, the comparison virgin bed V4 had adsorbed 58.1 g COD, with a breakthrough of 54.3%. Thus, the CRE for R4 was calculated to be 43.3%. This regeneration required 40.8 watt hours per gram of regenerated capacity.

Regeneration of Carbon Bed R5

Based on the performance of R4 after regeneration, it appeared that: (a) some chloride was desirable in the regeneration brine, and (b) applying a high cell potential (approximately 3 volts) or current (approximately 9.5 amperes), or a combination of the two during a regeneration was as detrimental to the carbon as too high a chloride level in the brine.

Therefore, another regeneration was performed on a spent carbon column. This column is designated R5.

The R5 regeneration was conducted at a reduced voltage (approximately 2.81 volts), which gave a cell current of 2.13 amperes. Although this regeneration procedure had a duration of 61.0 hours, the total energy input was only 365 watt hours or 4.70 watt hours per gram of adsorbed COD organics.

The B5 brine used in the regeneration of R5 was similar to B2, used in the 61.1 percent regeneration, except that the chloride level was not increased over that found in the tap water used to make the brine. The composition of brine B5 was 35.4 g/l Na_2SO_4, 87 mg/l chloride, and it had a pH of 8.0 and conductivity of 30,000 μmho/cm.

Bed R5 was rinsed with the previously determined optimum number of six changes of water. However, the time that each rinse was allowed to remain in the bed was decreased to 120 minutes. Conductivity measurements seem to indicate that this length of time is sufficient to remove enough of the remaining brine to prevent salt contamination of the wastewater when the bed is placed back on stream.

Bed R5 was returned to the wastewater plant and spent with 5,466 liters of wastewater in a 334 hour period. At the end of this period of time, the composite sample of effluent from R5 had a breakthrough dissolved COD value 86.9 percent of the dissolved COD of the common feed. It was calculated that R5 had, at this time, adsorbed 21.5 grams of dissolved COD material. At the same time, virgin comparison bed V5 saw 5,222 liters of wastewater. The last composite sample of effluent of this comparison bed contained material which exerted a COD 82.7 percent of the common feed wastewater. It was calculated that V5 adsorbed 49.2 grams of COD and, based on these loadings, that regeneration R5 had a CRE of 43.6 percent. The regeneration required 17.0 watt hours per gram of COD adsorption capacity regenerated.

It did not appear from the appearance of the B5 brine or from the final appearances of the rinses that bed R5 had been peptized or otherwise affected by the regeneration procedure. Yet its performance was still only about 44 percent of that of virgin carbon. Two possibilities exist to explain this poor regeneration result. The first is the possibility that the 61 hour duration of the R5 regeneration was too long. That is, the length of time that a carbon bed was subjected to a regeneration treatment affected in some way the resultant performance. It will be noted (Table 3) that the first regeneration (R1) in which the carbon was peptized, was of only 24.0 hours duration.

The second possibility is that bed R5 was not fully regenerated due to the lower power and energy input. That is, little adsorbed COD material was actually removed in the R5 regeneration.

Based on the assumption that the R5 regeneration was conducted for too long a period of time, the subsequent regenerations were conducted, as follows:-

Regeneration of Carbon Bed R6

The R6 regeneration was performed on the exhausted R5 bed. This was both the second attempt at a multiple regeneration and an attempt to improve on the performance of the R5 regeneration.

The R6 regeneration was conducted for 6.25 hours, at an average cell potential of 3.20 V. Cell current was 12.5 amp. Total energy input was thus 250 watt hours, i.e. 11.60 watt hours per gram COD adsorbed.

Following regeneration, bed R6 was rinsed in a similar manner as was R5.

Bed R6 and its virgin comparison bed V6 were then placed on stream for 704 hours. Bed R6 treated 10,551 liters of chlorinated sand filtered secondary effluent, adsorbed 36.6 g of dissolved COD organics, and produced a final composite effluent sample which had a COD breakthrough which was 88.8 percent of that of the untreated feed. Bed V6 treated 9,432 liters of wastewater, adsorbed 98.0 g of dissolved COD material, and after 704 hours of stream produced a composite effluent sample with a COD breakthrough that was 64.0 percent of that of the sample feed composite (Table 2).

Comparing the COD capacity of R6 and V6 (Table 1), it was calculated that the CRE of R6 was 37.4 percent. Comparing the two multiple regenerations (R6 and R3), it may be seen that the CRE of R6 is only 9.2 percent less than that of bed R3 (46.6 percent), but was obtained using only a 6.25 hour duration regeneration compared to a 70.0 hour duration regeneration in the case of R3 (Table 4). The 37.4 percent of regenerated capacity required only 250 total watt hours, or 6.82 watt hours per gram of COD capacity regenerated. This can be compared to the 1182 watt hours of 36.3 watt hours per gram required for the 46.6 percent regeneration of carbon bed R3 (Table 1, Figures 5 and 6).

As speculated, it is possible that the 61.0 hour regeneration procedure performed on R5 had already destroyed much of the COD adsorption capacity of this bed and therefore, in this second treatment, the 37.4 percent capacity was actually the maximum degree of regeneration obtainable.

Regeneration of Carbon Bed R7

Because of the indications that long regeneration times were deleterious, the last regeneration (R7) was of extremely short duration. R7 was performed on a carbon

bed that had been exhausted once (as the virgin comparison bed to R5). Regeneration R7 was conducted for 3.75 hours, at a cell potential of 3.20 volts. Average cell current was 13.3 amps. Total energy input was 160 watt hours (3.24 watt hours per gram COD adsorbed). Brine composition (R7) was 35.4 g/l Na_2SO_4 and 32 mg/l chloride, with a pH of 7.08 and a conductivity of 32,500 μmho/cm. Bed R7 was rinsed in a similar manner to bed R5.

After rinsing the regenerated bed, R7, was placed on stream (simultaneously with R6) and compared with virgin bed V6. Bed R7 was on stream for 663 hours and treated 9705 liters of wastewater. Virgin bed V6 was on stream 704 hours and treated 9432 liters of wastewater. The last composite sample of effluent from R7 had a COD that was 92.5 percent of that of the feed. Bed R7 adsorbed a total of 46.1 g of COD organics (Table 3).

By comparison with virgin bed V6, the CRE for regeneration R7 is calculated to be 47 percent. This degree of regenerated capacity required a total of 159.60 watt hours or 3.42 watt hours per gram of COD capacity regenerated (Table 1).

Other than the 61.1 percent regenerated capacity of bed R2, this 47.1 percent is to date the maximum amount of COD capacity restored by the electrochemical regeneration process after the first regeneration of a carbon bed. However, the 47.1 percent regeneration required only about one-quarter of the energy (3.42 versus 15.3 watt hours per gram of COD capacity regenerated), and only about one-seventh of the time required to obtain the 61.1 percent regenerated capacity (3.75 versus 30.5 hours).

General Discussion

The results of the six regenerations, R2-R7, have been plotted in Figures 5, 6, 7 and 8. (R1 has not been plotted because of the low CRE obtained.) Figure 5 shows CRE as a function of energy input per gram COD capacity regenerated. Figure 6 shows CRE as a function of total energy input. In Figure 7, CRE is shown as a function of chloride concentration in the brine used in the regeneration treatment. Figure 8 shows CRE as a function of the conductivity of the brine used in the regeneration treatment. Figure 9 shows COD loading on the carbon column as a function of the volume of the column, for each of the seven regenerated columns and six virgin comparison columns.

Considering first Figures 5-7, it can be seen that the CRE of all regenerations (with the exception of R1) was in the range 37% to 61%. The highest CRE

observed was 61%, for R2. There does not, however, appear to be any obvious relationship between CRE and the variables plotted. It is felt that this is due to the fact that in these regenerations several parameters were changed simultaneously. This was done in an attempt to rapidly determine optimum parameters, within the limited timespan of the program. Unfortunately, however, this approach has only obscured the nature of the results.

In Figure 9, loading curves are presented for carbon columns R2-R7 and V2-V6, showing COD loading as a function of effluent volume passed through the column. These curves point up another problem encountered in interpretation of the data from these experiments, namely variation in the performance of the virgin columns themselves. This variation may also have tended to obscure the results of the different regeneration procedures.

In conclusion however, it can be said that:

(1) Electrochemical carbon regeneration is capable of restoring at least 61% of the virgin column capacity.
(2) The nature of the variables affecting the carbon regeneration efficiency has not been sufficiently explored, i.e. it will be necessary to explore both a wider range of the variables studied to date, as well as additional variables, in order to optimize the regeneration process.
(3) The length of time required to exhaust the 600g carbon columns, and the variable nature of the effluent to which they were exposed make it difficult to perform the necessary number of parametric investigations.

Recommendations have been made for future extension of this work (Section II) in a manner to avoid the difficulties mentioned above.

The following section discusses some cost considerations relative to electrochemical carbon regeneration, which indicate the value of continuing this work.

SECTION VI

COMPARISON OF THE EFFICIENCY OF THERMAL AND ELECTROCHEMICAL REGENERATION, AND PRELIMINARY ESTIMATE OF OPERATING COST

The following discussion is presented to compare the presently used carbon regeneration procedure (thermal regeneration) with electrochemical regeneration, from the standpoint of efficiency and cost. Data on thermal regeneration is derived from the available literature (references 4 and 5); that on electrochemical regeneration from the results presented in Section V, for regenerations R2 and R3. Based on a report of the operation at Pomona, California (reference 4), the dissolved COD capacity of thermally regenerated carbon after the first regeneration is 79 percent of that of virgin carbon.

Based upon Environics' data presented above, the maximum first electrochemical regeneration efficiency was 61 percent. It can be shown then that electrochemical regeneration is approximately 77.2 percent as efficient as thermal regeneration. Further optimization of the electrochemical operating parameters could result in higher efficiencies for electrochemical regeneration.

After a regenerated column is spent at Pomona, it is thermally regenerated once again. The efficiency of this twice-regenerated column is reported to be 74% of a virgin column. However, if this number is corrected for biological activity (which, it is reported, accounts for 10 percent of the removal of COD organics), the net efficiency of a second thermal regeneration is 64 percent.

This, however, must also be corrected for the addition of 8.5 percent of virgin carbon to the regenerated bed, to restore the weight of carbon physically lost in removing the carbon from the contactor, transporting the carbon to a furnace, and actually regenerating the carbon. This 8.5 percent is believed to be the minimal loss encountered in the process.

Based upon the above data, the capability of a second regeneration by thermal means is calculated to be 62.5 percent.

Environics' data indicate that a second regeneration, using electrochemical means, as high as 42 percent has been obtained (in R3, described above). It is believed that in the full or pilot-scale operation of an electrochemical regeneration process at a water treatment plant, there would be only a small loss of carbon. Some carbon fines would be lost but only those smaller than the pore diameter of the screen used in the carbon contactors. In the present project the only loss of carbon occurred during regeneration through the upper unscreened plumbing of

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the contactor. This loss was estimated to be a maximum of 3 percent of the carbon weight and was probably much less.

The data obtained to date on electrochemical regeneration has not been sufficiently exhaustive to obtain the optimum results. It is believed that further small scale work, using the isotherm technique, possibly would demonstrate that electrochemical regeneration is as, or more effective than thermal regeneration.

Based upon the dosage of activated carbon per gallon of wastewater treated at Lake Tahoe (reference 5), one can project the electrical power cost of electrochemical regeneration using data obtained to date. Power requirements for a 61.1 percent regeneration (R2) are shown to be 16 watt hours per gram of COD capacity restored. At Lake Tahoe, the average COD is 20 mg/l and approximately 50 percent of this is removed across the activated carbon where a dosage of 207 pounds of carbon per million gallons of wastewater is employed (reference 5). Therefore, it would require 605.6 watt hours of treat 1000 gallons of water (0.207 pounds of carbon per 1000 gallons). At a power cost of one cent per kilo watt hour, this is 0.61 cents per 1000 gallons.

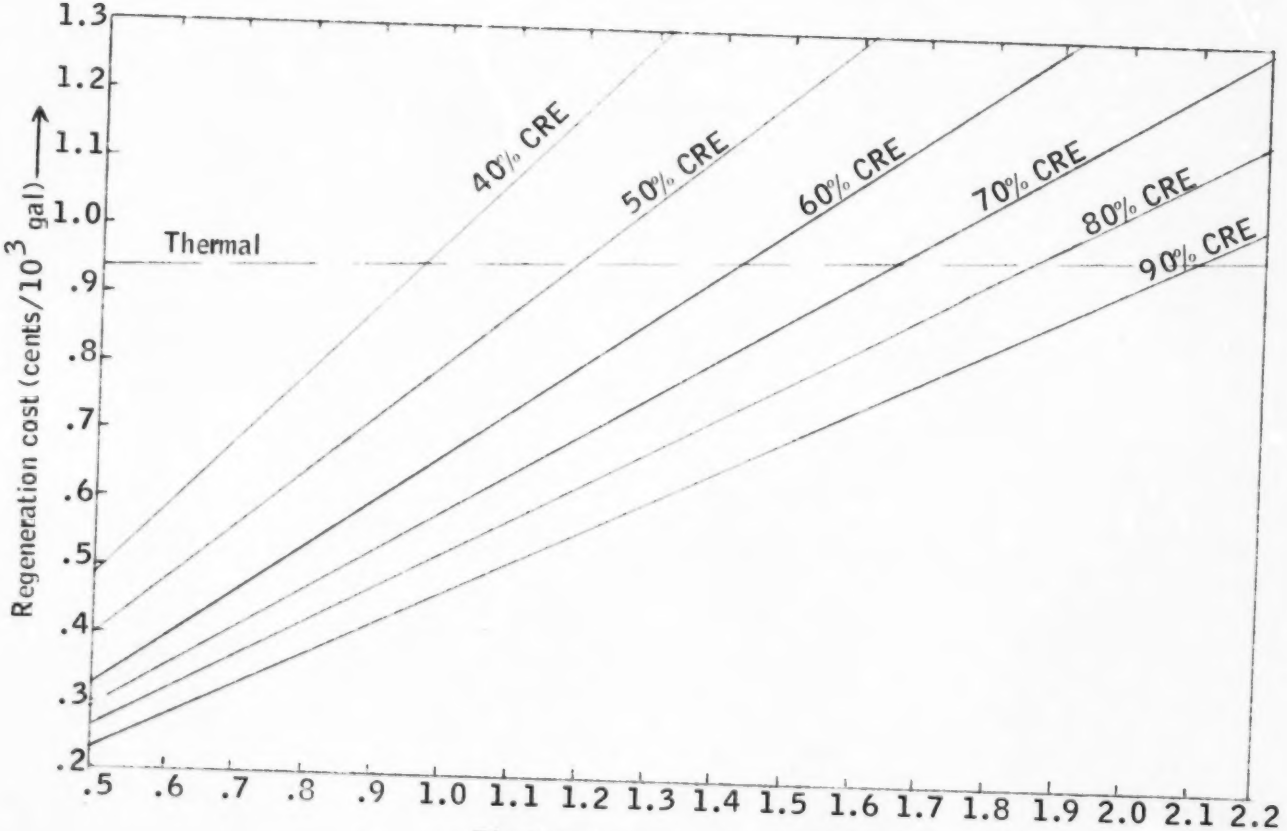
The data for circulation power has been obtained in the Environics' laboratory and showed that 25 watts are required to recirculate 5 liters per minute of electrolyte for 12 hours through 600 grams of carbon (the amount of carbon in an Environics' contactor). Thus, 47 watt hours would be required for 1000 gallons of Lake Tahoe-type wastewater and the electrical cost for circulation power is calculated to be 0.047 cents per 1000 gallons. Six one-liter changes of water are required for rinsing the brine from the carbon. It has been estimated that, in practice, this would be the same as pumping each of six one-liter rinses at two liters per minute through 600 grams of carbon, and that this would require ten watts of power. It has been further calculated that it would require 9.4 watt hours per 1000 gallons of Lake Tahoe-type wastewater to remove this brine. The total cost required for rinsing would then be 0.0094 cents per 1000 gallons of Lake Tahoe-type wastewater. Summation of the total power requirement is 662 watt hours, assuming the power cost is one cent per kilowatt hour; the total cost to treat 1000 gallons of Lake Tahoe-type wastewater using electrochemical techniques is 0.66 cents. This is equivalent to 3.2 cents per pound of carbon at a dosage rate of 207 pounds per million gallons. This figure could be reduced if the power requirement for a more complete regeneration (greater than 61.1 percent) is actually less than 16 watthours per gram of COD capacity restored.

Similar calculations on the cost of electrochemical regeneration have been performed, assuming different values for power cost and CRE obtainable. Figure 10 shows the results of these calculations, i.e. electrochemical regeneration cost

(in cents/1000 gallons wastewater treated) is shown as a function of power cost, for different values of carbon regeneration efficiency. Also shown is the cost of thermal regeneration, based on data from reference 5. Electrochemical regeneration costs include costs for electrolysis, pumping and rinsing. Both costs are shown exclusive of labor.

It can be seen from this figure that electrochemical regeneration is less costly than thermal regeneration, even at a CRE of 40%, if power can be purchased for one cent/kilowatt hour. If power cost is 1.5 cents/ kw hr, a CRE of 60-65% is required for break-even.

Furthermore, both costs as shown in Figure 10 are exclusive of labor. Labor costs for thermal regeneration more than double the regeneration cost for this process. Since it is expected that labor cost for electrochemical regeneration will be much lower than for thermal, the tradeoff is even more favorable toward electrochemical regeneration than shown.



Regeneration Cost Vs Power Cost

Table 10

SECTION VII

ACKNOWLEDGEMENTS

Appreciation is expressed to the Irvine Ranch Water District for the use of their water reclamation plant, and especially to Mr. Ray L. Stoyer, Director of Planning, and to Mr. James L. Peoples, Chief Plant Operator, for their assistance in conducting our field tests at the reclamation plant.

The donation of materials from the Calgon Corporation and the valuable help from Mr. Lowell W. Arnold of the Water Management Division is gratefully acknowledged. The operation and maintenance of the laboratory work, was performed in part by Mr. Arthur S. Lucero, who also served as Assistant Principal Investigator. Mr. Y. C. Lee contributed as Contract Program Manager. Mr. M. L. Stary aided in the preparation of the final report.

This project was initiated at McDonnell-Douglas Astropower Laboratory under Federal Water Pollution Control Administration Contract 14-12-839. We acknowledge the many helpful discussions with Dr. G. A. Guter and Mr. H.K. Bishop.

The support of the project by the Environmental Protection Agency made this work possible. The help and patience of Mr. Charles E. Myers, EPA Project Officer for this contract is especially appreciated.

SECTION VIII

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